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ORGANIC EMISSIONS EVALUATION OF A PAINT BAKE OVEN WITH CATALYTIC INCINERATION

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16. ABSTRACT

This report describes a test program conducted at a paint bake oven facility. The purpose was to measure total hydrocarbon (THC) concentrations at the inlet and outlet of an incinerator with heat recovery used to reduce organic solvent emissions. Data were also collected to evaluate the energy efficiency and economics of the system compared to other THC control alternatives.

The incinerator system was designed by Schweitzer Industrial of Madison Heights, Wisconsin, and incorporates DuPont's Torvex catalyst with platinum to enhance hydrocarbon reduction in the process stream. The incinerator system includes a heat exchanger following the catalyst bed. The heat exchanger "effectiveness" of this configuration was 82 percent, allowing for a recovery of 35.1 percent of total thermal energy from the gas stream.

Analysis of the annualized costs of thermal and catalytic incineration and carbon adsorption was performed. Because the concentration of hydrocarbons to the control device is small, the annualized cost of carbon adsorption is less than other control devices.

Measurements conducted at the inlet and outlet of the incinerator indicate an average reduction in emissions of approximately 86 percent. Bypassing the incinerator with a fraction of the total gas stream resulted in an emissions reduction to the atmosphere of 70 percent.

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FOREWORD

When energy and material resources are extracted, processed, converted, and used, the related pollutional impacts on our environment and even on our health often require that new and increasingly more efficient pollution control methods be used. The Industrial Environmental Research Laboratory-Cincinnati (IERL-Ci) assists in developing and demonstrating new and improved methodologies that will meet these needs both efficiently and economically.

This report documents a recently completed project that accessed a catalytic incineration system with heat recovery to reduce organic solvent emissions from small parts automobile paint bake ovens. The findings of this report primarily document the emissions reduction and heat recovery potentials of this system. In addition, an assessment of the economic feasibility of incineration with heat recovery is presented. For further information, contact the Nonferrous Metals and Minerals Branch, IERL, Cincinnati, Ohio.

David G. Stephan
Director
Industrial Environmental Research Laboratory
Cincinnati, Ohio

ABSTRACT

This report describes sampling methods and results of a field test program conducted at the Mack Truck, Inc. paint bake oven facility located in Allentown, Pennsylvania. The purpose of the test program was to measure total hydrocarbon (THC) concentrations at the inlet and outlet of an incinerator with heat recovery used to reduce organic solvent emissions. Data were also collected to evaluate the energy efficiency and economics of the system compared to other THC control alternatives.

The incinerator system was designed by Schweitzer Industrial of Madison Heights, Wisconsin, and incorporates DuPont's Torvex catalyst with platinum to enhance hydrocarbon reduction in the process stream. The incinerator fuel is No. 2 distillate oil injected through a Model 500, combination overpack gas/oil burner manufactured by Maxon of Muncie, Indiana. The incinerator system includes a heat exchanger following the catalyst bed. The gas stream being heated is circulated to the E-coat oven, thus replacing a direct heat source otherwise required. The heat exchanger "effectiveness" of this configuration was 82 percent, allowing for a recovery of 35.1 percent of total thermal energy from the gas stream.

An analysis of the annualized costs of thermal and catalytic incineration and carbon adsorption was performed. Because the concentration of hydrocarbons to the control device is small, the annualized cost of carbon adsorption is less than other control devices.

Measurements conducted at the inlet and outlet of the incinerator indicate an average reduction in organic emissions of approximately 86 percent. Bypassing the incinerator with a fraction of the total gas stream resulted in an emissions reduction to the atmosphere of 70 percent.

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LIST OF SYMBOLS

C_p = heat capacity, joules/g-mole ^oC or cal/g-mole ^oC (Btu/lb mole ^oF)

Δh = enthalpy, cal/gm (Btu/lb)

$\dot{M}$ = mass flowrate, g-mole/min (lb-mole/min)

MW = molecular weight g/g-mole (lb/lb-mole)

$\dot{q}$ = overall heat transfer rate or cal/min (Btu/min)

$\dot{Q}$ = energy, joules/hour or cal/min (Btu/min)

T = temperature, ^oC (^oF)

$\dot{W}$ = mach joules/hr, cal/min (Btu/min)

X_i = volume fraction

LIST OF METRIC-TO-ENGLISH CONVERSION FACTORS

To convert from	To	Multiply by
Calories (gm)	Btu	0.0039683
Calories (gm)	Joules	4.184
Calories (gm)	kW-hours	1.162222×10^{-6}
Cubic meters	cubic feet	35.3107
Joules	Btu	9.4709×10^{-4}
Joules	calories	0.239
Kilograms	pounds	2.2046226
Meters	feet	3.2808399
Pascals	atmospheres	9.869×10^{-6}
Pascals	inches of water ($4^{o}C$)	4.015×10^{-3}
Pascals	pounds per square inch	0.000145

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The authors gratefully recognize the technical contributions of the following Acurex staff members: Mr. Mark Chips for his assistance and technical support during the field test program, and Mr. Robert Larkin for his review and comments.

SECTION 1

INTRODUCTION

This report describes the test methods and results of a program conducted on a catalytic incinerator system designed to control hydrocarbon emissions from small parts paint bake ovens. The objective of this program was to evaluate the organic solvent emissions reduction potential of a catalytic incineration system with heat recovery. Selection of the facility to be tested was made by the Nonferrous Metals and Minerals Branch, IERL-Ci. Testing was conducted April 14 through 18, 1980 by Acurex Corporation at the Mack Truck, Inc. facility in Allentown, Pennsylvania.

Three 1-hour tests to measure THC at the inlet and outlet of the incineration system were conducted. Other sampling locations included a bypass stream around the incinerator and ambient air. Gas phase conditions at each test location were measured using accepted EPA methods. Gas phase conditions were also determined for the inlet and outlet streams from the heat exchanger following the incinerator. Process parameters were monitored for the test period using existing Mack Truck, Inc. instrumentation. Data was also collected to evaluate the economic feasibility of the incineration system.

SECTION 2

CONCLUSIONS

The catalytic incineration system tested at Mack Truck, Inc. was effective in reducing organic emissions from the small parts bake ovens. Based on the average emission rates calculated for the three 1-hour tests, the incinerator reduced organic emissions by 86 percent. Bypassing the incinerator with a fraction of the total gas stream resulted in an emissions reduction to the atmosphere of 70 percent.

The effectiveness of the heat exchanger, used to heat air for the E-coat oven, was 82 percent, allowing for a recovery of 35.1 percent of the total thermal energy from the gas stream entering the heat exchanger thus allowing for reduced energy requirements. Larger fuel savings may be possible to Mack Truck, Inc. if the combustion air and gas stream being incinerated are preheated (rather than the E-coat oven air) since more energy can be transferred to these streams, thus allowing reduced fuel requirements. A further analysis of the heat exchanger showed that 41.5 percent of the energy required to operate the incineration system could be recovered. This is equivalent to 63 kg/hr of the no. 2 oil being burned.

The primary disadvantage of thermal or catalytic incineration is the cost associated with the fuel requirements. Carbon adsorption has lower operating costs than the incinerators but higher capital costs. All of these processes can reach near total control of the THC emissions. The major difference between incineration and adsorption was the annualized cost of each. The annualized cost for carbon adsorption was $34,100 per year. The corresponding costs for thermal and catalytic incinceration with heat recovery are $200,000 and $77,500 per year, respectively. These cost differentials are due to the low concentration of THC being controlled, thus allowing no reduction in the fuel requirements for incineration (due to the combustible content of the gas), or requiring negligible regeneration costs for carbon adsorption.

SECTION 3

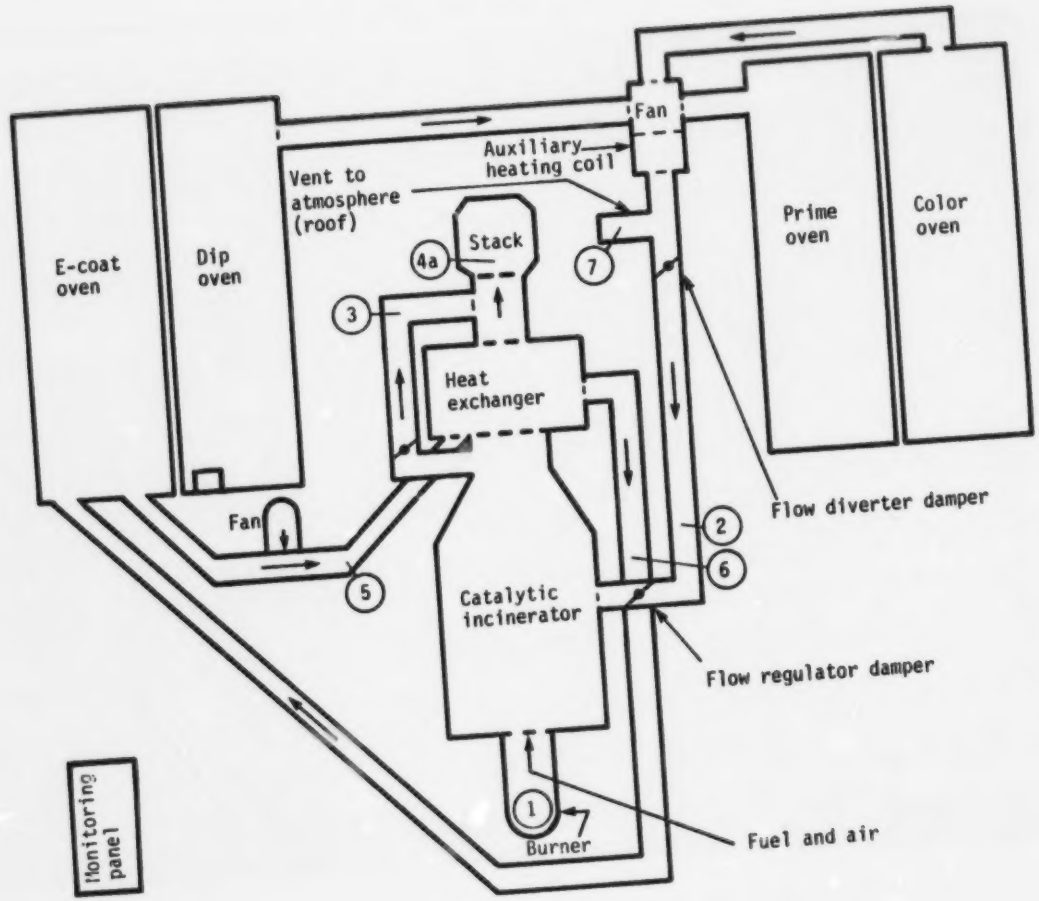
PROCESS DESCRIPTION AND SYSTEM OPERATIONS

The incineration system, located in building 514 at the Mack Truck, Inc. facility in Allentown, Pennsylvania, was designed by Schweitzer Industrial, Madison Heights, Wisconsin in 1976. The purpose for installation was to control odorous emissions from small parts paint bake ovens located in the building. The system incinerates off-gases from three paint bake ovens (prime oven, color oven, and dip-coat oven), and supplies heat via a heat exchange system to the E-coat (electrodeposition) oven. Figure 1 presents a schematic illustration of the system. The system is normally operated continuously during the first and second shifts 6 days per week. The identification numbers of each test location are shown in Figure 1 with a description of each presented in Section 4.

The incinerator is fired using no. 2 distillate fuel oil in a Model 500 combination overpack gas/oil burner manufactured by Maxon of Muncie, Indiana. Fuel consumption during normal operation is approximately 170 liters/hour.

DuPont's Torvex catalyst with platinum is incorporated in the incinerator to enhance the unit's efficiency for emission reduction. The unit was tested using the catalyst bed installed in December 1976. The useful life of the catalyst as installed is 5 years. Therefore, this test program was conducted when less than 20 percent of the catalyst life remained, assuming the absence of any "blinding," (i.e., plugging of the catalyst pores). The catalyst is scheduled to be changed (regenerated) during the 1980 calendar year.

The incineration system includes a heat exchanger serving to heat the off-gas (air) from the E-coat oven for subsequent return to that oven. The heat exchanger, which follows the catalyst bed, also includes a bypass system as shown in Figure 1. The bypass is the means for regulating the temperature of the gas going to the E-coat oven.



○ = sample locations

Figure 1. Schematic of the Mack Truck, Inc. incineration system.

SECTION 4

SAMPLING AND ANALYSIS PROCEDURES

This section describes the procedures used to measure the incinerator operating parameters and THC emissions.

4.1 DETERMINATION OF GAS PHASE CONDITIONS

Prior to measuring THC concentrations, measurements to determine the gas phase condition at each sampling location were conducted. These activities included:

- The location of the sampling site (existing) and number of sampling points to be used
- The velocity, temperature, pressure, and volumetric flowrate of the gases at the sampling sites
- The molecular weight and moisture content of the gases to be sampled

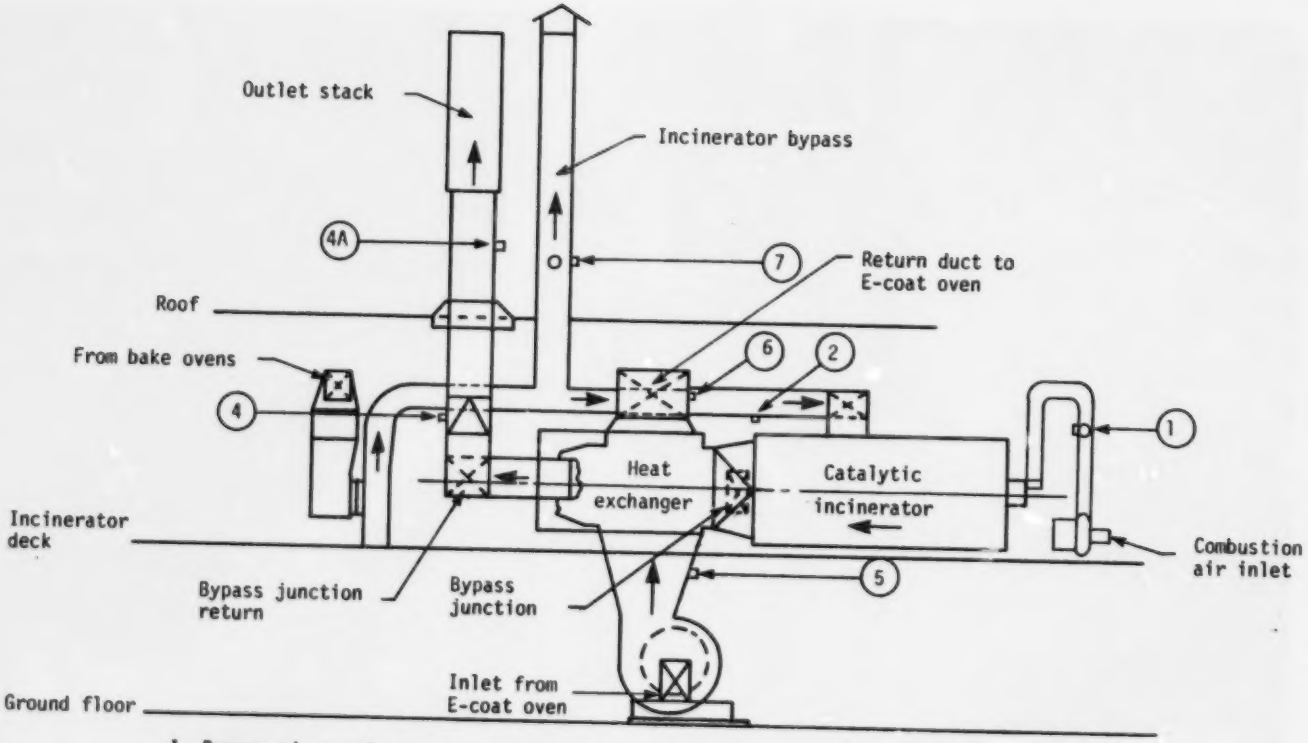
These measurements were performed using EPA Methods 1 through 4 (reference 1) identified as follows:

- Method 1 -- Sample and Velocity Traverse for Stationary Sources
- Method 2 -- Determination of Stack Gas Velocity and Volumetric Flowrate (S-Type Pitot Tube)
- Method 3 -- Gas Analysis for Carbon Dioxide, Oxygen, Excess Air, and Dry Molecular Weight
- Method 4 -- Determination of Moisture Content in Stack Gases

This data was collected at each of the appropriate sampling locations. Table 1 presents a summary of the measurements conducted at each location.

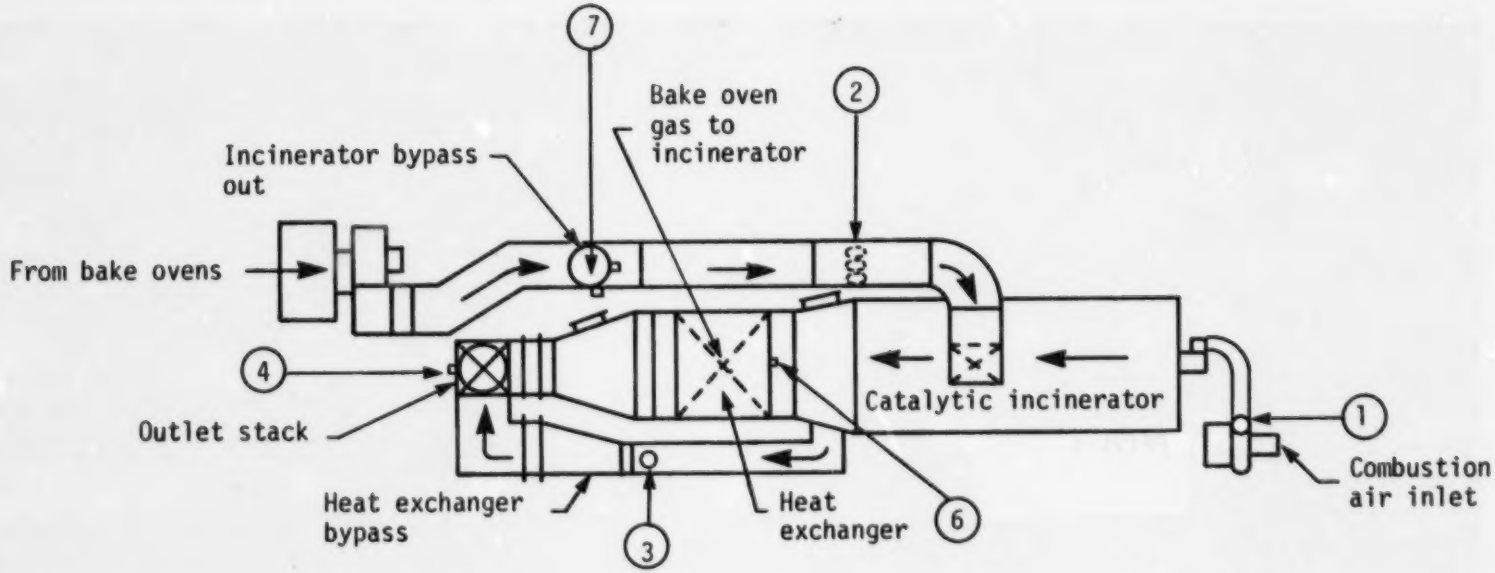
4.1.1 Location of Sampling Points

Figures 2 and 3 present schematics (side view and top view) of the incinerator system showing the location and description of sampling port locations. (The sampling locations are also illustrated schematically in Figure 10.) Each location and the measurements conducted at each are



1 Burner air supply
2 Paint bake oven off-gas inlet to incinerator
3 Heat exchange bypass (see Figure 3)
4 Incinerator exhaust gas outlet (THC measurement)
4A Incinerator exhaust gas outlet (velocity measurement)
5 E-coat oven exhaust makeup air inlet to heat exchanger
6 Heat exchange outlet (return to E-coat oven)
7 Paint bake oven incinerator bypass (velocity measurement)

Figure 2. Side view of incineration system showing sampling and/or measurement locations.



1 Burner air supply
2 Paint bakeoven off-gas inlet to incinerator
3 Heat exchange bypass
4 Incinerator exhaust gas outlet (THC measurements)
4A Incinerator exhaust gas outlet (see Figure 2)
5 E-coat oven exhaust and makeup air inlet to heat exchanger (not shown)
6 Heat exchanger outlet (return to E-coat oven)
7 Paint bakeoven incinerator bypass (roof)

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Figure 3. Top view of incineration system showing sampling and/or measurement locations.

TABLE 1. GAS CONDITION DATA ACQUISITION SUMMARY BY SAMPLE LOCATION AND METHOD

Location	Method 1: Traverse	Method 2: Velocity/ temperature profile	Method 3: Molecular weight	Method 4: Moisture	THC
1 Burner air	X	X	b	d	X[e]
2 Paint bake oven off-gas (inlet to incinerator)	X	X	b	d	X
3 Heat exchanger bypass	X	X	X	f	
4 Incinerator exhaust gas outlet (incinerator deck)			X	X	X
4A Incinerator exhaust gas outlet (roof)	X	X	c	c	
5 E-Coat oven exhaust gas outlet	X	X	b	d	
6 Heat exchanger outlet		a	b	a	
7 Paint bake oven incinerator bypass (roof)	X	X	b	b	g

[a]Same as location 5
[b]Assumed to be that of air since no combustion products were present
[c]Same as location 4
[d]Determined using psychrometric methods
[e]Monitored periodically
[f]Determined stoichiometrically (computer)
[g]Same as location 2

described in the following paragraphs.

Location 1: Burner Air Supply--

Burner air supply sampling ports were located 1.2 diameters upstream from a 90-degree bend in the ductwork and 4.4 diameters downstream from the blower fan flange coupling on a vertical 0.25-meter duct. Using EPA Method 1 criteria, the maximum number of sampling points (10) were chosen and two traverses were performed at right angles to one another.

Location 2: Paint Bake Oven Off-Gas (Inlet to Incinerator)--

This location was sampled at three ports located on the underside of a rectangular duct, 5.7 equivalent diameters downstream from the nearest disturbance and 1.4 equivalent diameters upstream from a 90-degree bend leading to the incinerator. The internal dimensions of the duct were 0.74 meter wide by 0.53 meter high. Nine sampling points (centroid area) were chosen using EPA Method 1.

Location 3: Heat Exchanger Bypass--

Due to the inacessibility of the bypass duct location and the hazardous (extreme heat) conditions at this location, only one sample port was used to access the bypass gas stream. This port was located in the duct 5.6 equivalent diameters downstream from a bend in the duct and 0.9 equivalent diameters upstream from a transition leading to the incinerator exhaust stack. The internal dimensions of the duct were 0.36 meter wide by 1.17 meters high. A total of 15 traverse points were located in a vertical plane 0.008 meter apart.

Location 4: Incinerator Exhaust Gas Outlet (Incinerator Deck)--

There were no velocity or temperature profile data collected at this location due to a port nipple assembly too small for access, and due to its proximity of the return of the bypass gas, less than 1 meter upstream of location 4.

Location 4A: Incinerator Exhaust Gas Outlet (Roof)--

One existing 0.10-meter port was located on the roof 6.9 diameters upstream from a transition in the incinerator unit and 6.0 diameters from the stack exit. Using EPA Method 1, the maximum number of sampling points (24) were chosen and one traverse for stack gas velocity and temperature was performed.

Location 5: E-Coat Oven Exhaust and Makeup Air (Inlet to Heat Exchanger)--

This duct provides a widening transition from the fan housing to the incinerator heat exchanger and does not conform to recommended EPA Method 1 sampling location criteria. However, due to the inaccessibility of other areas, the port in the transition ductwork leading to the heat exchanger was tested. The internal dimensions of the duct at this plane were calculated to be 1.17 by 1.32 meters. A total of 23 traverse points were located in a horizontal plane 0.05 meter apart.

Location 6: Heat Exchanger Outlet (Return to E-Coat Oven)--
One existing port, located at a duct transition, was traversed for temperature profile only. A total of 12 traverse points were located on a horizontal plane approximately 0.008 meter apart.

Location 7: Paint Bake Oven Incinerator Bypass (Roof)--
This duct is used to vent bake oven off-gas to the atmosphere if the incineration system is not able to accept all or a fraction of the oven off-gas. Velocity and temperature profiles were conducted by performing a single traverse at the existing port located on the roof. The maximum number of sample points (24) were used to traverse the 0.66-meter diameter duct.

4.1.2 Determination of Gas Velocity and Volumetric Flowrate

Gas velocity and volumetric flowrate data were obtained using EPA Method 2. An S-type pitot tube was used to traverse each test location. Velocity and static pressures were obtained using a 0- to 0.025-meter inclined scale manometer manufactured by Dwyer Instruments Inc. of Anaheim, California. A leak check of the system was performed before and after each traverse to ensure that the system was leak-free.

Gas temperatures were measured concurrently with the pressure measurements (for the velocity traverse) using a type K (chromel-alumel) thermocouple in near proximity to the pitot tube. A Minimite potentiometer manufactured by Thermo Electric Corporation of Saddle Brook New Jersey was used to measure the thermocouple signal.

4.1.3 Determination of Stack Gas Composition

Samples of the gases at locations 3 and 4 were collected using a stainless steel probe and a one-way squeeze bulb to transport the gases to a standard Orsat Flue Gas Analyzer manufactured by Hayes Republic of Michigan City, Indiana. The percent CO_2, O_2, and CO were determined volumetrically using EPA Method 3. A minimum of three runs were performed at each location sampled and the results were averaged to determine the stack gas composition. As noted in Table 1, the other gas streams were assumed to be the same as atmospheric air.

4.1.4 Determination of Stack Gas Moisture Content

Stack gas moisture content was determined at locations 3, 4, and 4A using a computer program and assuming a typical fuel analysis for the fuel oil and other data obtained from the gas composition measurement.

Stack gas moisture content at locations 2 and 3 was determined using EPA Method 4. However, the low moisture content of the gas at location 2 produced no measureable amount of condensation and was therefore undeterminable. To ascertain the moisture content of the gas at location 2, psychrometric data was obtained and the percent moisture calculated using the vapor pressure of water at the corresponding wet bulb temperature

(see Appendix for calculations). The psychrometric method of moisture determinations was also used to determine the moisture content of the burner air (location 1).

4.2 TOTAL HYDROCARBON MEASUREMENTS

Hydrocarbon concentrations in the gas stream were determined using a THC analyzer. The following paragraphs describe the procedures followed for each test sequence.

4.2.1 Total Hycrocarbon Measurement System Description

A Model 400 Hydrocarbon Analyzer, manufactured by Beckman Instruments, Inc. of Fullerton, California, was used to measure THC in the sampled gas streams. This analyzer uses the flame ionization detection method (FID).

The FID was operated using zero grade 1.0 hydrogen fuel and zero air supplied by Airco Industrial Gases of Riverton, New Jersey. Hydrogen fuel and zero air pressure were set at 207 and 103 kPa, respectively, using internal differential pressure regulators in the analyzer.

The sampling system is shown in Figure 4. The gas sample was extracted from the stack via a 7-micron sintered stainless steel filter, SS-4FE-7, manufactured by Nupro Valve Company of Willoughby, Ohio. The filter served to remove fine particulates which could, if allowed to pass into the THC analyzer, occlude the FID sample inlet capillary. A 0.006-meter O.D. stainless steel probe connected the filter unit to the heated sampling line via a three-way, 0.006-meter stainless steel ball valve. This valve allows the introduction of sample gas or calibration gas depending on which mode of operation is desired. A 6.1-meter, heat-traced 0.001-meter Teflon sample line manufactured by Technical Heaters, Inc. of San Fernando, California was used to transport the gas sample to the vacuum pump. The temperature of the sample lines was maintained at 394°K by separate proportional temperature controllers supplied by the manufacturer. At this point, the sample lines were connected to another three-way valve which was coupled to the vacuum side of a Model 917CA18, single-stage, Teflon-coated diaphragm vacuum pump, manufactured by Thomas Industries of Sheboygan, Wisconsin. The three-way valve at this location allowed the instrument operator to switch the sample lines intermittently between sample locations. From the vacuum pump, the sample was transported to an isothermal well constructed from approximately 12.2 meters of 0.006-meter O.D. annealed copper tubing. This assembly was maintained at 450°K by use of a proportional temperature controller manufactured by RFL Industries, Inc. of Boonton, New Jersey and a 2.4-meter 1,000-watt Glascol Heating Tape. The isothermal well ensured that the sample gas or calibration gas routed to the FID was maintained at a constant temperature to eliminate possible instrument drift (due to temperature changes) during sampling and calibration procedures. It also provided the system with some degree of independence from process changes at the sampling location.

From the isothermal well, the sample gas was transported directly to the FID. Sample pressure from the vacuum pump outlet was maintained at 31 kPa by an internal pressure regulator/bypass system contained in the THC analyzer. This pressure setting provided a sample flowrate through the system

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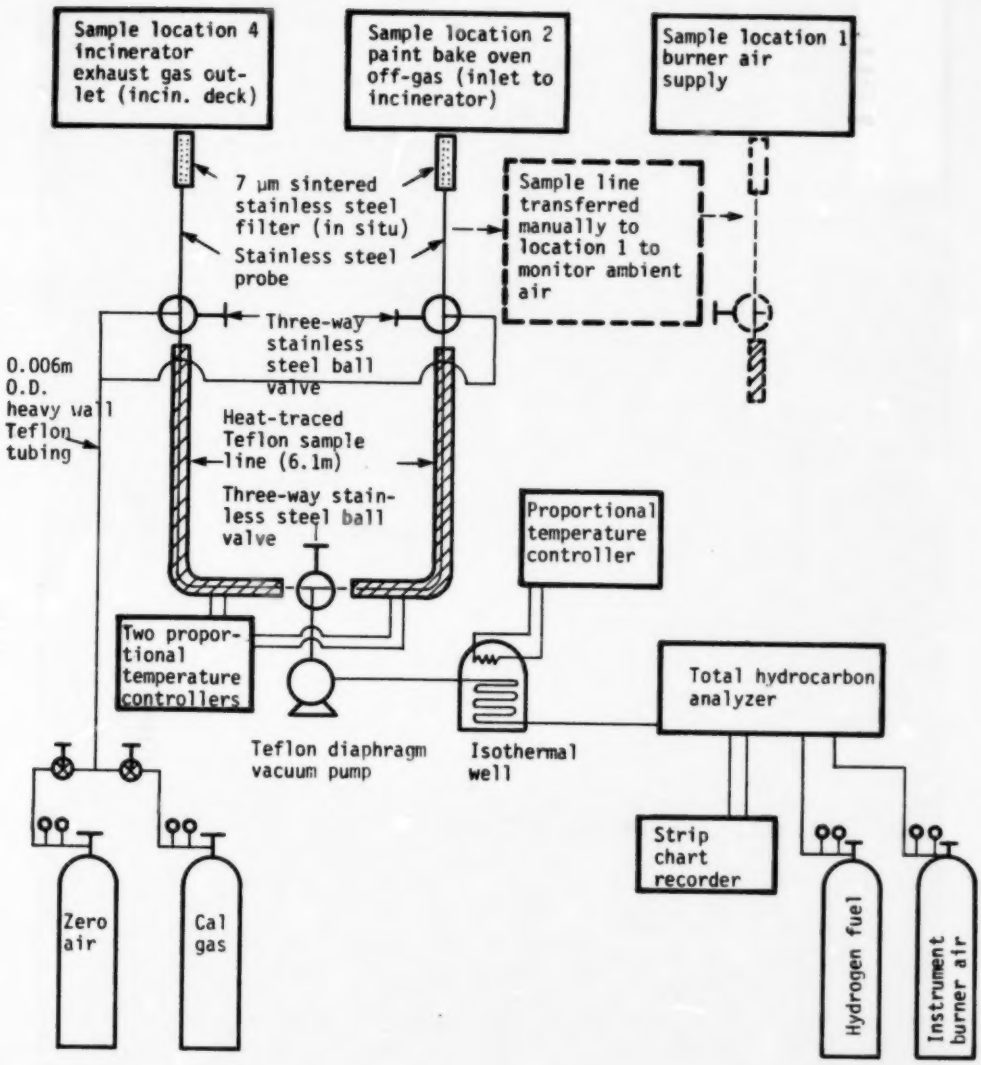
Sample location 4 incinerator exhaust gas outlet (incin. deck)
Sample location 2 paint bake oven off-gas (inlet to incinerator)
Sample location 1 burner air supply
7 μm sintered stainless steel filter (in situ)
Stainless steel probe
Sample line transferred manually to location 1 to monitor ambient air
Three-way stainless steel ball valve
0.006m O.D. heavy wall Teflon tubing
Heat-traced Teflon sample line (6.1m)
Three-way stainless steel ball valve
Proportional temperature controller
Two proportional temperature controllers
Total hydrocarbon analyzer
Teflon diaphragm vacuum pump
Isothermal well
Strip chart recorder
Zero air
Cal gas
Hydrogen fuel
Instrument burner air

Figure 4. Schematic of total hydrocarbon sampling system.

of approximately 3.0 liters per minute with excess sample vented to the atmosphere.

4.2.2 Total Hydrocarbon System Calibration and Operation

Prior to operation and calibration, the sampling system was operated at approximately 422°K and was purged for several hours with nitrogen to remove any traces of residual hydrocarbon contamination in the lines. During this "bake-out" procedure, stainless steel tube unions, filters, and probes were also heated using a propane torch. Prior to, and at the completion of testing, a leak test was performed on the sampling system by sealing the inlet to both sampling lines and assuring that the external (THC analyzer) system sample pressure dropped to zero.

Before and after each test series the THC analyzer was calibrated using zero air and a mixture of 535 ppm methane in nitrogen. All calibration gases were transported from their respective cylinders to the three-way stainless steel ball valve located between the 0.006-meter stainless steel probe and heat-traced, 0.01-meter Teflon sample line via heavy wall 0.006-meter Teflon tubing. To calibrate the analyzer, the three-way valve was positioned to block the sample probe and filter and allow the calibration gas to pass into the heat traced sample line. By introducing the calibration components at this location, the sample gases and calibration gases were treated in the same manner and possible undesirable effects due to absorption (wall loss) in the sampling line and system were avoided.

4.2.3 Total Hydrocarbon Sampling Method

Each test series consisted of approximately 60 minutes of continuous monitoring in 10-minute intervals at location 2 (incineration inlet), and location 4 (incinerator outlet). Background (ambient) hydrocarbon levels were monitored periodically at the burner air supply (location 1). The heat exchanger bypass (location 3) was also monitored to determine variations in background hydrocarbon levels.

Though THC concentrations were expected to be independent of the velocity profile, stratification of hydrocarbons in the process streams at sampling locations 2 and 4 were monitored by conducting a sampling traverse at each location. Traversing was conducted using the same sample point locations previously described.

SECTION 5

DATA PRESENTATION

Data from the test program provided information on gas phase condition and composition, hydrocarbon concentrations, and atmospheric emissions. This section of the report presents the data collected during the field test program.

5.1 GAS CONDITION AND COMPOSITION

Table 2 presents a summary of the collected and calculated gas condition and composition data at the various locations defined in Figure 1. These data will be used to calculate the mass flowrates of hydrocarbons, incinerator efficiencies, heat recoveries, and other pertinent information.

5.2 TOTAL HYDROCARBON EMISSIONS

THC was recorded as equivalent methane at the inlet and outlet of the incinerator and in the surrounding atmosphere (background). Table 3 presents a summary of the measurement results. The high and low concentration data presented represents the minimum or maximum concentration detected during the test period. Average values presented in the table were calculated by reading the THC concentration on the output for each run and averaging the data over the total sample time at that location. Each data point, therefore, represents a 10-minute sampling period, with an average concentration calculated from 1-minute incremental positions on the recorder output. A total of eight data points for the incinerator inlet and nine data points for the incinerator output are presented in Figure 5. This figure shows the randomness of the THC concentrations in the gas stream entering the incinerator. Data was not available from Mack Truck, Inc. to indicate if the decreasing concentrations during each sampling period were due to a batch drying sequence. Parts are randomly placed in the ovens as they are painted. In addition, the concentrations would also be sensitive to the size of each batch of parts being dried. The data in Figure 5 show that the outlet of the incinerator is insensitive to fluctuations in the inlet concentration. This may indicate that, for this catalytic system, the surface reactions are rate controlling rather than the diffusion or adsorption/desorption steps.

Ambient measurements of the THC concentration in the heat exchanger bypass duct and combustion air stream averaged 18 ppm during the tests. These data points are also presented in Figure 5.

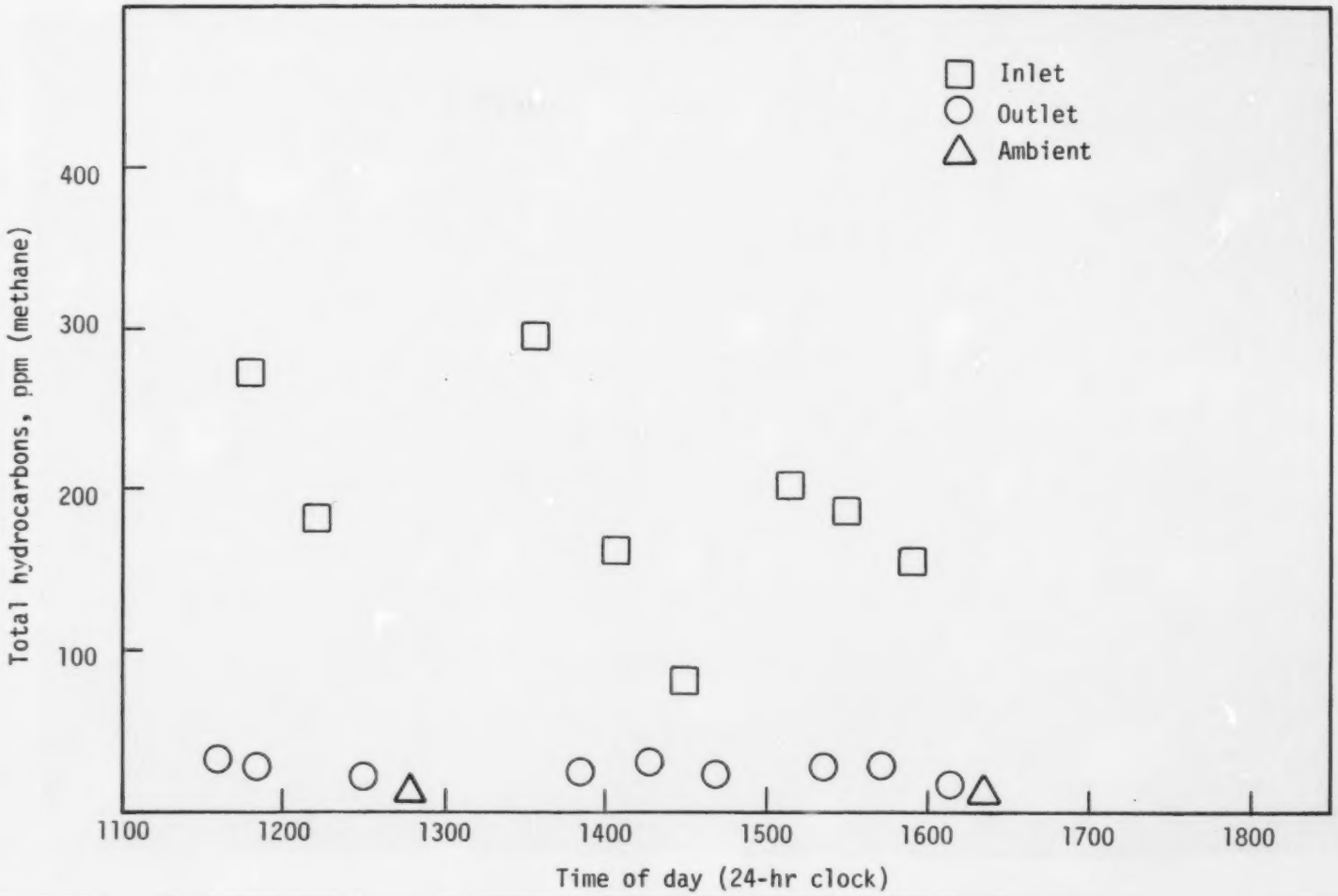

Figure 5. Graph of inlet and outlet THC concentration versus time for April 16, 1980.

TABLE 2. SUMMARY OF GAS CONDITION DATA AND CALCULATED RESULTS

Location description	Duct dimensions (meters)	Duct area (m^2)	Barometric pressure (kPa)	Static pressure (duct) (kPa)	Absolute pressure (duct) (kPa)	Average gas temperature (^oK)	Average velocity pressure (kPa)	Average gas velocity (m/sec)	Volumetric flowrate actual (m^3/min)	Volumetric flowrate standard[e] (m^3/min)	% H_2O	M_s molecular weight wet (g/g-mole)
1. Burner supply air	0.254	0.05	98.54	+4.73	103.3	316	0.059	15.91	47.86	44.83	1.03	28.85
2. Paint bake oven off-gas (inlet to incinerator)	0.74 W x 0.53 H	0.39	98.54	+1.84	100.4	411	0.039	12.37	291.63	197.17	4.6	28.45
3. Heat exchanger bypass	0.36 W x 1.17 H	0.42	98.54	+0.22	98.77	777	0.029	12.71	316.28	113.54	b	b
4. Incinerator exhaust gas outlet (incinerator deck)	N/A	N/A	-	-	-	-	-	-	-	-	2.9[c]	28.56
4A. Incinerator exhaust outlet (roof)	0.76	0.46	99.75	-0.11	99.65	631	0.045	17.43	477.17	212.28	b	b
5. E-coat oven exhaust and makeup air (inlet to heat exchange)	1.17 W x 1.32 H	1.54	98.54	+1.99	100.56	444	0.034	11.15	1,031.87	670.91	1.03	28.85
6. Heat exchange outlet[a] (return to E-Coat oven)	N/A	N/A	-	-	-	495	-	-	-	d	d	d
7. Paint bake oven incinerator bypass (roof)	0.66	0.34	98.54	+0.62	99.14	396	0.105	3.21	65.97	45.68	4.6	28.45
8. Fuel composition									2.37[f]			

[a]Temperature profile only
[b]Same as location 4a
[c]Calculated using typical no. 2 fuel oil analysis (computer)
[d]Same as location 5
[e]Standard conditions are 20°C (293°K), and 29.92 inches of mercury (101.3 kPa), and zero moisture
[f]Actual mass flow, kg/min
N/A -- Not applicable

TABLE 3. SUMMARY OF THC MEASUREMENT RESULTS

Item	Test times (4-16-80) (24-hr clock)	Test duration (minute)	THC concentration, ppm (methane)								
			Location 2			Location 4			Ambient		
			High	Low	Average	High	Low	Average	High	Low	Average
Test 1	1132 to 1236	64	283	157	225	37	18	27	20	15	17.5
Test 2	1326 to 1446	80	359	61	179	37	20	26	--	--	--
Test 3	1503 to 1615	72	231	137	180	29	10	23	23	13	18
Average	--	72	291	118	195	34	16	25	22	14	18

To assess the independence of THC concentration with respect to gas velocity over the cross-section of the sampling location, a traverse of each test location was conducted. The THC concentration at each traverse point (centroid areas previously determined for velocity/temperature profile) was measured. A 10-ppm THC concentration change from point to point was noted during 90-second sampling intervals.

5.3 HEAT EXCHANGER PERFORMANCE

To determine heat exchanger performance, mass and energy balances were developed using the flow system shown in Figure 9 and data from Table 2. It should be noted that the heat exchanger bypass stream joins the heat exchanger outlet flow prior to the measurement point, location 4A; thus, the actual exit conditions (i.e., flowrate and temperature) must be computed. Also, since the molecular weights shown in Table 2 are very close to that of air (i.e., very dilute streams), the flow properties will be assumed as those of air (e.g., standard density is 1.201 kg/m^3). Based on this assumption, the following balance is determined:

Based on this assumption and the measured volumetric flowrates given in Table 2:

Flow inflow	=	2.37 kg/min
Air inflow	=	53.8 kg/min
Off-gas inflow, location 2	=	237 kg/min
Exchanger bypass flow	=	136 kg/min
Incinerator exhaust outlet, location 4A	=	255 kg/min

a mass balance yields the heat exchanger inflow as follows:

Exchanger inflow = fuel + air + off-gas - bypass
157 = 2.37 + 53.8 + 237 - 136

Therefore, the heat exchanger inflow is 157 kg/min. A mass balance on the heat exchanger outlet:

Exchanger outlet = incinerator exhaust - bypass
119 = 255 - 136

yields an exchanger outlet mass flux of 119 kg/min.

These balances reveal a significant discrepancy between the computed in- and outflows for the exchanger. To determine the source of this discrepancy, the sample point locations, traversing technique, and profile results were examined. As a result, it appears that the velocity profiles and resulting flowrates for locations 3 (Figure 6) and 4A (Figure 7) are high due to flow stratification. The air inflow is also high but the flow at location 2 (Figure 8) is the most reliable. Consequently, the remaining computations will be done using the inlet flow determined.

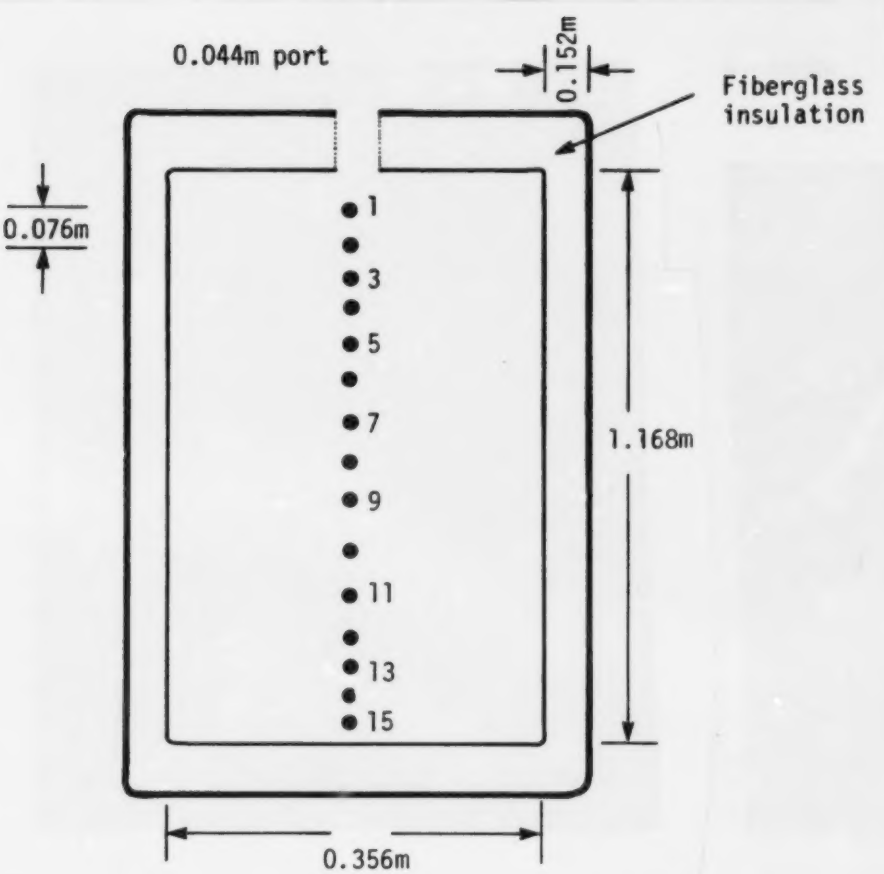

Figure 6. Schematic showing cross-section of sampling location 3 (heat exchanger bypass) indicating sample traverse locations.

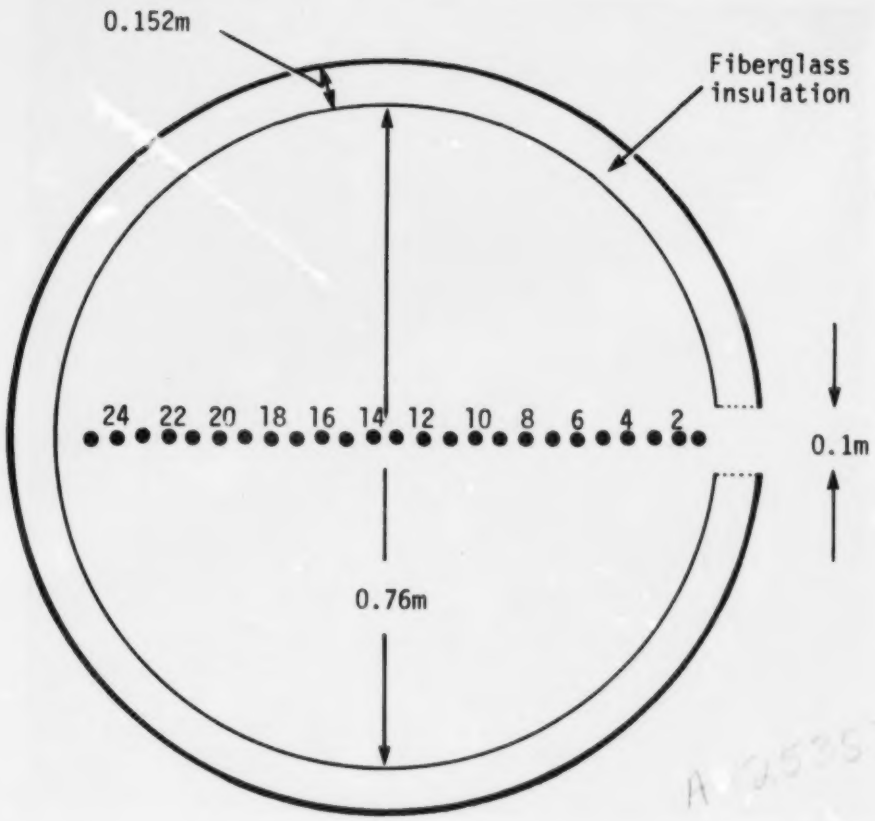

Figure 7. Schematic showing cross-section of sampling location 4a (incinerator outlet duct) indicating sample traverse locations.

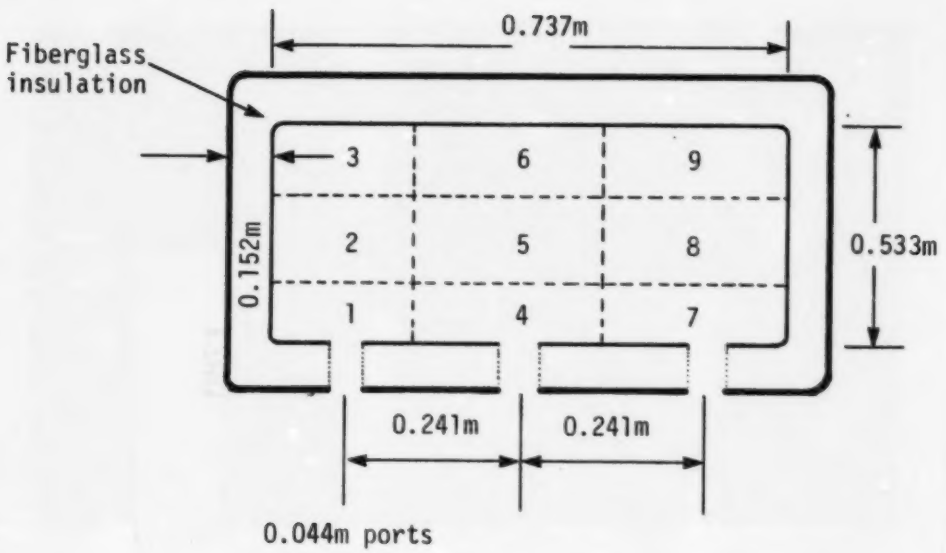

Figure 8. Schematic cross-section of sampling location 2 (paint bake oven off-gas (inlet to incinerator)) indicating sampling traverse locations.

The rate of energy recovered from the off-gas in the heat exchanger is computed from the following equation:

$$\dot{q} = \dot{M} C_p (T_{h_{in}} - T_{h_{out}}) \quad (1)$$

where:

$\dot{q}$ = Overall heat transfer rate

$\dot{M}$ = Mass flowrate

C_p = Heat capacity

$T_{h_{in}}$ = Inlet temperature, hot-side

$T_{h_{out}}$ = Outlet temperature, hot-side

The heat exchanger outlet temperature ($T_{h_{out}}$), was not measured; however, it can be computed based on the temperature and flowrate for stream 3, the temperature for stream 4A, and the total flow based on inlet measurements as follows:

$$\dot{M}_3 T_3 + \dot{M}_4 T_{h_{out}} = \dot{M}_{4A} T_{4A}$$

where

$$\dot{M}_{4A} = \dot{M}_{fuel} + \dot{M}_{fuel} + \dot{M}_{air} + \dot{M}_2$$

or numerically:

$$\dot{M}_{4A} = 2.37 + 53.8 + 237 = 293 \text{ kg/min}$$

and

$$T_{h_{out}} = \frac{(293\ (631) - (136)\ (777)}{157} = 504^{\circ}K$$

Note that the unreliability in volumetric flow measurements for streams 3 and 4A would make this value a maximum.

The resulting heat recovery rate based on this computed outlet temperature and the assumption that the temperature of the gas entering the heat exchanger is equal to the bypass temperature is then 0.740×10^6 joules/hr The total energy flowrate for the heat exchanger inlet stream is:

$$\dot{M} C_p T_{h_{in}}$$

or 2.09×10^6 joules/hr (based on a value of 0.241 cal/gm ^oK for C_p), thus 35.1 percent of the energy flux is recovered in the exchanger. To complete the energy balance, the cold-side recovery is also determined. Using equation 1 applied to the cold-side, the heat transferred is computed to be 10.1×10^6 cal/min. The difference could be attributed to losses and/or data uncertainties.

The performance of a heat exchanger is a function of the specific heats of both fluids, heat exchanger surface area, heat exchanger flow geometry, heat exchanger material thermal conductivity and fluid flowrates. This performance can be quantitatively expressed in terms of the "effectiveness" of the heat exchanger in transferring a given amount of heat (reference 2). Effectiveness can be defined as:

$$\varepsilon = \frac{\text{actual heat transfer rate}}{\text{maximum possible heat transfer rate}} \qquad (2)$$

The maximum possible heat transfer ($\dot{q}_{max}$) can be computed using the equation:

$$\dot{q}_{MAX} = (\dot{M}C_p)_{MIN} \times \Delta T_{MAX} \qquad (3)$$

where:

$(\dot{M}C_p)_{min}$ = The fluid with the highest potential for maximum heat transfer rate based on minimum mass flowrate and heat capacity.

ΔT_{MAX} = The maximum thermal driving force

For this case, the maximum thermal driving force (ΔT_{MAX}) is:

$$\Delta T_{h_{in}} - T_{C_{in}}$$

where

$T_{C_{in}}$ is the temperature of the cold-side gas entering the heat exchanger.

Therefore, equation 3 becomes:

$$q_{MAX} = (\dot{M}C_p)_{MIN} \times (T_{h_{in}} - T_{C_{in}}) \quad (4)$$

The actual heat transfer rate was obtained previously. Since for this case, $M_h C_{p_h}$ is equal to $(M C_p)_{MIN}$, equation 2 can be expressed as:

$$\varepsilon = \frac{T_{h_{in}} - T_{h_{out}}}{T_{h_{in}} - T_{C_{in}}} \quad (5)$$

Substituting the available data (Table 2) into equation 5, the transfer effectiveness (ε) is equal to 82 percent (based on the values determined using inflow data).

SECTION 6

DATA ANALYSIS

This section presents the final results of the test program. The information presented includes the emission reduction capability of the incinerator and the overall control effectiveness, and the fuel savings which result from waste heat recovery.

6.1 INCINERATOR EMISSION CONTROL EFFECTIVENESS

A summary of the emission rates for each incinerator sample location is presented in Table 4. Emission rates presented in the table were calculated using the average THC concentrations (Table 3) for all three test runs and the measured volumetric flowrate (Table 2) for that location. Also presented is an emission rate for the paint bake oven incinerator bypass (location 7). Since this bypass was vented directly to the atmosphere from the paint bake oven off-gas duct, it affects the overall control effectiveness. The average THC concentration at location 2 was used to calculate the gaseous emission rate to the incinerator and the incinerator emission control effectiveness.

The THC concentrations measured at the inlet of the incinerator (location 2) fluctuated widely as shown in Figure 5. These fluctuations were due to the semicontinuous operation of the paint spray booths and bake ovens. As a batch of newly coated parts entered the oven (at random), solvent emissions increased. In addition, the concentration of THC in the gas stream was also impacted by the size and number of parts in each batch.

Fluctuations in THC concentration at the outlet of the incinerator (location 4A) also occurred. Using the overall test averages for THC measured at the inlet and outlet of the incinerator, the incinerator emission control effectiveness in reducing THC was 86 percent. The overall effectiveness of the system in reducing emissions to the atmosphere, including the bypass, was 70 percent.

6.2 OVERALL SYSTEM ENERGY BALANCE

To operate the THC emission control system on an automobile paint bake oven, certain additional energy requirements result for the Mack Truck, Inc. facility, the combustor fuel energy and the energy expended in the fans and blowers needed to deliver the combustor air and overcome additional pressure drop. For this system, however, some of this additional, control-related energy is recovered for use in heating the E-coat oven. In this section an overall system energy balance will be developed. Based on this balance, the

TABLE 4. SUMMARY OF EMISSION RATES

Sample location	Description	Emission rate	
		Kg/hr (as methane)	PPM (as methane)
2	Paint bake oven off-gas (inlet to incinerator)	1.552	195
4A	Incinerator exhaust gas (outlet of incinerator)	0.214	25
7	Paint bake oven incinerator bypass	0.360	195
1	Burner air	0.0325	18

effectiveness of the energy recovery system (i.e., the heat exchanger) will be determined as a percent of the control energy penalty.

An energy balance is developed for the control volume outlined in Figures 9 and 10. This control volume does not include the work input to the oven fan or alternate heat source for the E-coat oven since these are assumed to be in operation for both the controlled and uncontrolled systems.

For energy conservation, the energy inflow minus energy outflow is equal to energy accumulation. For steady-state operation, the energy accumulation is zero. Therefore:

$$\dot{Q}_{air} + \dot{Q}_{fuel} + \dot{Q}_{gas} + \dot{W}_{work} - \dot{Q}_{exchange} - \dot{Q}_{stack} - \dot{Q}_{losses} = 0 \quad (6)$$

where $\dot{W}_{work}$ is the energy input of the heat exchanger fans and blowers. The work input from the fans will be taken at maximum rating.

$$\dot{W}_{burner\ fan} = 11{,}186 \text{ watts} = 40.27 \times 10^6 \text{ joules/hr}$$

$$\dot{W}_{exchanger\ fan} = 74{,}570 \text{ watts} = 268.5 \times 10^6 \text{ joules/hr}$$

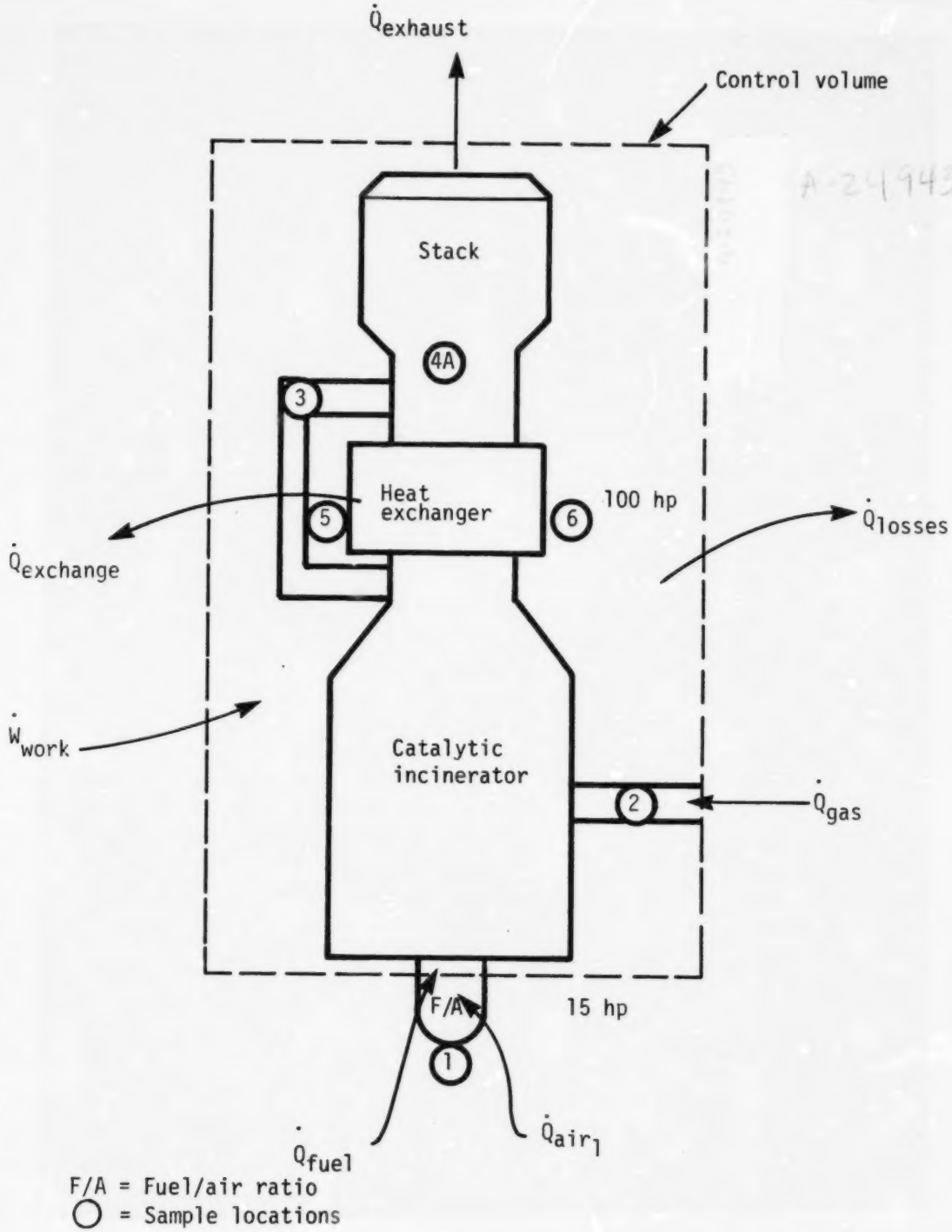

Figure 9. Control volume for heat balance.

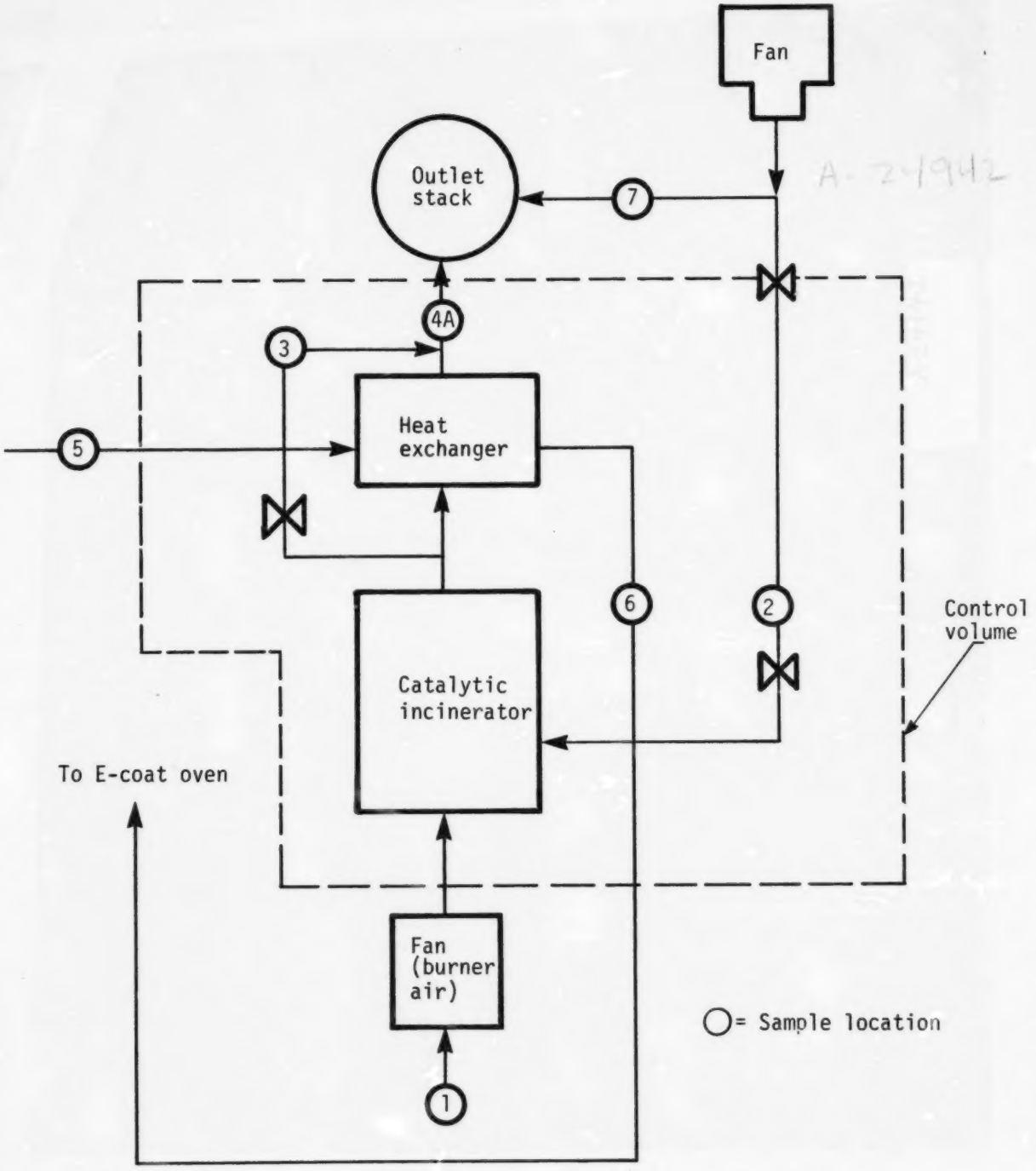

Figure 10. System schematic.

therefore,

$$\dot{W}_{work,\ max} = 308.77 \times 10^6 \text{ joules/hr}$$

The heat exchange rate determined in the previous section is:

$$\dot{Q}_{exchange} = 2{,}637 \times 10^6 \text{ joules/hr}$$

The burner is firing No. 2 fuel oil at 166.5 liters/hr or 142.3 kg/hr, with a heating value of 41,869 kJ/kg. Therefore:

$$\dot{Q}_{fuel} = 142.3 \text{ kg/hr} \times 41{,}869 \text{ kJ/kg} = 5{,}958 \times 10^6 \text{ joules/hr}$$

To determine the energy input from the gas streams, it is necessary to determine the enthalpy of the gas stream based on the volume fraction of each component. The streams were assumed to be perfect gas mixtures (i.e., the mole fraction is equal to the volume fraction, and the specific heats are constant in this temperature range). Therefore, the energy content of each remaining stream becomes:

$$\dot{Q}_j = \left[\Sigma x_i \, \overline{C}_{p_i} \, T_j - T_o + \Delta h_{H_2O} + \Delta h_R\right] \frac{\dot{M}_j}{(MW)_j}$$

where

$\dot{Q}_j$ = energy flux of stream j

$\dot{M}_j$ = mass flux of stream j

$(MW)_j$ = molecular weight of stream j

x_i = volume fraction of component i

$\overline{C}_{p_i}$ = mean heat capacity of component i

T_j = temperature of stream j

Δh_R = enthalpy of reaction for CO to CO_2 and trace THC oxidation

T_o = ambient temperature

and $$\Delta h_{H_2O} = \dot{M}_j \, x_{H_2O} \left[\bar{C}_{p_l} \left(T_{D.P.} - T_o \right) + \bar{C}_{p_v} \left(T_j - T_{D.P.} \right) + \Delta h_r \right]$$

where $\bar{C}_{p_l}$ = heat capacity of H_2O liquid

$\bar{C}_{p_v}$ = heat capacity of H_2O vapor

$T_{D.P.}$ = dew point temperature

Δh_v = heat of vaporization of H_2O at dew point temperature

Based on the above, the remaining heat fluxes are as follows:

$\dot{Q}_{air} = 82 \times 10^6$ joules/hour

$\dot{Q}_{exhaust} = 5,864 \times 10^6$ joules/hour

$\dot{Q}_{gas} = 2,731 \times 10^6$ joules/hour

The energy balance is summarized in Table 5 and a sample calculation is presented in the Appendix.

As indicated, the imposition of THC emission control requires additional energy, namely, $\dot{Q}_{fuel}$, $\dot{Q}_{air}$ and $\dot{W}_{work}$. A portion of this energy is recovered to the E-coat oven by use of a waste heat exchanger prior to the stack. A measure of the effectiveness of this heat recovery system (η_c) is the

percent recovery of the additional energy required for the control device, (equation 8). This is given by:

$$\eta_c = \frac{\dot{Q}_{exchange}}{\dot{Q}_{air} + \dot{Q}_{fuel} + \dot{W}_{work}} \quad (8)$$

Note that $\dot{Q}_{gas}$ is not included in this energy term since the temperature of the gas from the other bake ovens to the incinerator is below that of the E-coat oven stream and that gas stream would exist for both controlled and uncontrolled systems. The recovery efficiency defined above is 41.5 percent. If the $\dot{Q}_{gas}$ were included to give an overall recovery efficiency η_o, as shown in equation 9, the value would be 32.7 percent.

$$\eta_o = \frac{\dot{Q}_{exchange}}{\dot{Q}_{air} + \dot{Q}_{fuel} + \dot{W}_{work} + \dot{Q}_{gas}} \quad (9)$$

The amount of heat recovered to the E-coat oven is equivalent to 63 kg/hr of No. 2 oil being burned.

TABLE 5. OVERALL SYSTEM ENERGY BALANCE (joules/hr)

Energy In:	
$\dot{Q}_{gas}$	$2,731 \times 10^6$
$\dot{Q}_{fuel}$	$5,958 \times 10^6$
$\dot{Q}_{air}$	82×10^6
$\dot{W}_{work}$	309×10^6
Total	$9,080 \times 10^6$
Energy Out:	
$\dot{Q}_{exhaust}$	$5,864 \times 10^6$
$\dot{Q}_{exchange}$	$2,637 \times 10^6$
$\dot{Q}_{losses}$	579×10^6[a]
Total	$9,080 \times 10^6$

[a]Determined by difference (6.4 percent losses)

SECTION 7

PROCESS ANALYSIS

The THC emission control effectiveness for the catalytic incinerator tested was 86 percent. Based on the installation date of December 1976 and a typical useful life of the catalyst of 5 years, it is estimated that less than 20 percent of the catalyst life remained for reaction at the time of these tests. This assumes that no particulate matter was present in the gas stream, which could blind the catalyst. The catalyst is expected to be operating such that the THC removal efficiency is near the maximum possible by this system.

An advantage of the incineration system is that it is insensitive to fluctuations in inlet concentrations (see Figure 7) and to changes in the types of contaminants entering the system. Again, this assumes that the species entering the incinerator will not blind the catalyst or result in the generation of a particulate matter that will require subsequent control.

The primary disadvantage to this control technique is the energy (fuel) required to raise the gas temperature to the level required for effective catalyst performance. Though not significant at the Mack Truck, Inc. installation due to the low concentration of organics, the greater the combustible content of the gas stream entering the system, the greater the offset of the energy requirement.

Thermal incineration (noncatalytic) depends on the thermal oxidation at temperatures in excess of 1500°F and residence times of 0.5 to 1.5 seconds (reference 3). Thermal incineration requires higher energy input to reach the operating temperature and must be constructed of heat resistant materials.

A reduction in this energy requirement can be obtained by preheating the gas stream entering the incinerator. Preheating the gas can be accomplished by means of heat exchangers or hot gas recirculation (combining a fraction of the incinerator effluent gas with the incinerator inlet stream). A typical heat recovery of 35 percent (reference 3) is used as the basis for determining the capital cost of each incineration system. This value is similar to the 35.1 percent heat recovery of the Mack Truck, Inc. system (see Section 5).

Either the thermal or catalytic incineration system is applicable to the control of hydrocarbons contained in a gas stream. The selection of this control technology by any facility over carbon adsorption is expected to be the economic burden of the energy requirements and the potential use of the solvent being controlled. Replacement parts (such as the catalyst) are

assumed to be provided as needed and therefore are not included. The use of an incinerator precludes the recovery of the organic species in the gas stream.

A carbon adsorption system consists of a bed of activated carbon capable of adsorbing organic species from a gas stream. The applicability (or at least the efficiency) of carbon adsorption is impacted by the organic constituents. Light organics, such as methane, are not effectively controlled by carbon adsorption. The carbon can be regenerated for reuse and has a useful life of approximately 5 years (reference 3). Typically, an adsorption system consists of two beds, one adsorbing and one being regenerated. The energy requirement is low, compared to incineration, and consists of the generation of steam for the regeneration step and cooling water (excluding gas moving equipment and instrumentation). If the operator elects not to recover the organics, a water effluent containing the organics would be generated, which may require treatment and/or disposal.

Carbon adsorption can result in near total hydrocarbon control. However, this system is adversely affected by variations in the types and concentrations of the contaminants present. Therefore, in applications similar to the Mack Truck, Inc. facility, the use of a carbon adsorption system should be based on the results from other similar applications presently in operation.

SECTION 8

COST ANALYSIS

The following discussion presents a generic cost analysis of each of the control devices discussed in Section 7. For the types of equipment under consideration, the number of operators and maintenance personnel and the useful life of the equipment and parts (catalysts and adsorbents) are equal (reference 3). Therefore, the primary differences between each system are the utility and capital charges, and the capital cost of each.

8.1 THERMAL AND CATALYTIC INCINERATION

The capital cost of a catalytic incineration system consisting of the catalyst beds, preheat burners, ductwork, fan and controls, is shown in Table 6. The addition of a heat exchanger for 35 percent heat recovery increases the capital cost by approximately 25 percent. The capital cost for a thermal incineration system is also shown in Table 6. Again, the addition of a heat exchanger for 35 percent heat recovery increases the capital cost by approximately 25 percent.

The operating cost differential between each pollution control device is impacted by the utility costs, capital recovery costs, and other indirect operating costs (taxes, insurance, administration). These costs are calculated using the following estimating procedures (reference 3):

Electricity	\$0.044/kWh (December 1977)
Fuel oil	\$0.13/liter (December 1977)
Steam	\$11.33/mKg (December 1977)
Indirect operating costs	4 percent of capital costs
Capital recovery cost	0.1102 based on 10 percent interest and a useful equipment life of 25 years

The major operating costs for incineration are presented in Tables 8 and 9. The annual fan power is calculated using the formula:

$$kWh = \frac{0.746\ (ACFM)\ (\Delta P)\ (SG)\ (H)}{6{,}356\ n} = 0.746\ (hp)(H) \qquad (10)$$

TABLE 6. CAPITAL COSTS FOR INCINERATORS (Reference 3)

	Capital cost[a]	Installed cost[b]
Thermal incinerator	69,000	111,100
Thermal incinerator with 35 percent primary heat recovery	86,000	138,500
Catalytic incinerator	82,000	132,000
Catalytic incinerator with 35 percent primary heat recovery	103,000	165,800

[a]December 1977 dollars
[b]Multiplier 1.61 (Reference 3)

TABLE 7. ANNUALIZED COSTS FOR INCINERATION WITHOUT HEAT RECOVERY (December 1977 dollars) (Reference 3)

	Thermal	Catalytic
Utilities (fans)	1,200	3,500
(fuel)	329,800	81,700
Indirect operating costs	4,400	5,300
Capital recovery	12,200	14,500
Total	347,600	105,000

TABLE 8. ANNUALIZED COSTS FOR INCINERATION WITH HEAT RECOVERY
(December 1977 dollars) (Reference 3)

	Thermal	Catalytic
Utilities (fans)	3,500	5,800
(fuel)	175,700	46,800
Indirect operating costs	5,500	6,600
Capital recovery	15,300	18,300
Total	200,000	77,500

where ACFM = Actual volumetric flowrate, Am^3/second

P = Pressure loss, kPa

n = Efficiency, assume 0.65

H = Hours of operation, assume 8,000 hr

SG = Specific gravity of air at operating temperature, gm/cm^3

The fuel use rate is estimated from EPA report 45/3-76-031 (reference 4). The fuel use is based on a 148.9°C inlet gas temperature with and without heat recovery.

Two systems are being considered here. The base system utilizes the recovered heat to heat the E-coat oven, thus reducing a direct fuel requirement. The second system, as shown in Figure 11, also recovers heat for preheating the gas being incinerated. For the configuration shown in Figure 11, if the heat exchanger efficiency is constant at 82 percent, the energy flux recovered by this configuration is 38.6 percent of the total energy available. Therefore, the configuration utilized by Mack Truck, Inc. may not utilize the available thermal energy in the most efficient manner. Preheating the gas could allow for a greater reduction in the fuel for the incinerator than would otherwise be required for heating the E-coat oven. This is due to the potential to extract more heat per unit area of heat transfer surface because of the larger temperature differential and would eliminate the need for a bypass stream.

Mack Truck, Inc. estimates that the installed cost of the catalytic incineration system was $180,000 in 1976. For purposes of comparing the economic feasibility, the difference between this cost and those presented in

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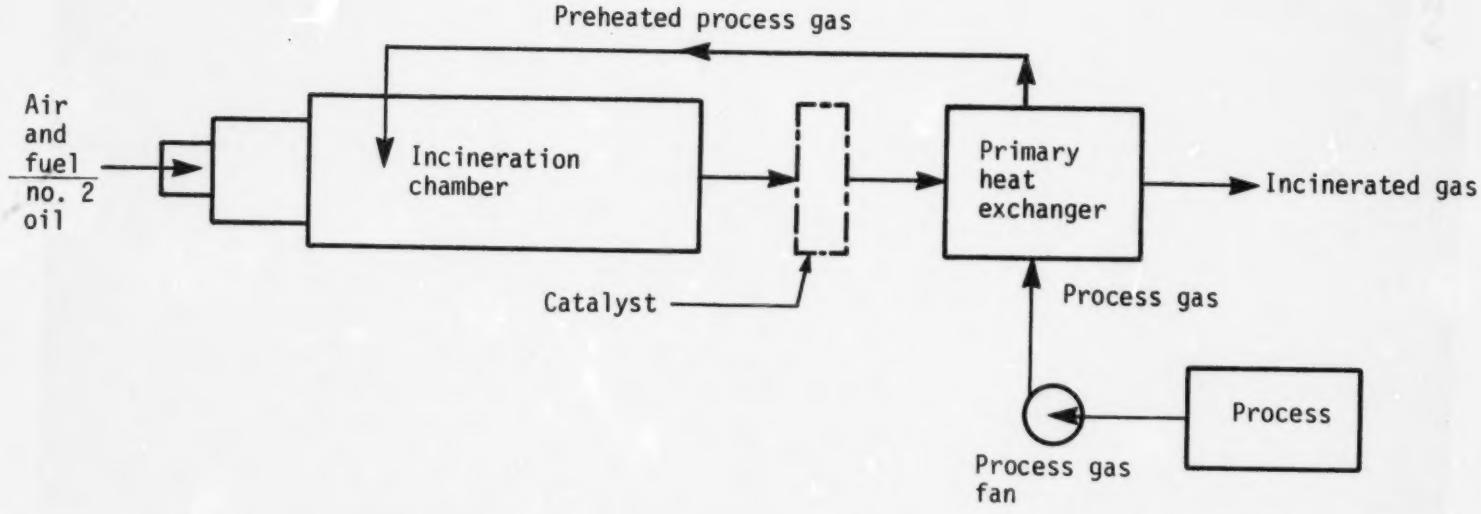

Figure 11. Primary heat recovery with catalyst.

Table 6 are assumed to be the auxiliary equipment (ducts, fans, insulation) necessary to heat the E-coat oven. In December 1977 dollars (6.5 percent escalation in equipment costs per year), this represents an installed capital cost of $38,400.

8.2 CARBON ADSORPTION

The capital cost of a carbon adsorption device is based on the total weight of carbon required for efficient control of the organic material being controlled and the configuration of the system. The carbon adsorption system used for this analysis includes a dual adsorber bed (one in the adsorption cycle and one in the regeneration cycle), fans, controls, and the steam generator. Using an average solvent recovery of 8 percent per 100 pounds of carbon requires 2,760 pounds of carbon per bed to reach 90 percent organic control effectiveness. This requires a capital investment of $114,400 for an industrial unit (reference 3). The total installed costs are $184,200.

The operating costs are impacted by each of the following (reference 1):

Steam consumption -- Assume 4 lb/lb of pollutant removed
Cooling water -- 12 gal/100 lb of steam
Electricity -- 5 hp/1,000 ACFM

Based on these assumptions, the annualized costs shown in Table 9 can be generated. It should be noted that concentration of pollutants being removed is small. Therefore, the operating costs are negligible with respect to those associated with higher pollutant concentrations. Though carbon adsorption appears more attractive than incineration for the Mack Truck, Inc. system, it may be economically unattractive in applications with high pollutant concentrations. In addition, it may be possible for even smaller annualized costs for Mack Truck, Inc., since they operate on a two-shift/day cycle. This would allow them to use one catalyst bed and regenerate during the third shift. Elimination of an adsorption bed would reduce the capital costs significantly.

TABLE 9. ANNUALIZED COSTS FOR CARBON ADSORPTION

Utilities	
Steam[a]	Negligible
Water[b]	Negligible
Fans[a]	9,200
Indirect operating costs[a]	4,600
Capital recovery[a]	20,300
Total	34,100

[a]See Section 8.1
[b]Cooling water $0.10/1,000 lb

REFERENCES

1. Federal Register, Vol. 42, No. 160, Thursday, August 18, 1977

2. Keith, F. Principles of Heat Transfer, International Textbook Co., 1965.

3. Neveril, R. B. Capital and Operating Costs of Selected Air Pollution Control Systems EPA-68-02-2899, U.S. Environmental Protection Agency, Research Triangle Park, North Carolina, December 1978.

4. C-E Air Preheater Co. Report of Fuel Requirements, Capital Cost and Operating Expense for Catalytic and Thermal Afterburners EPA-68-02-1473, U.S. Environmental Protection Agency, Research Triangle Park, North Carolina, September 1976.

APPENDIX A

PSYCHOMETRIC MOISTURE DETERMINATION CALCULATIONS - LOCATIONS 1 and 2

SAMPLE CALCULATION OF Q; FOR INCINERATION INLET STREAM

PSYCHROMETRIC MOISTURE DETERMINATION -- LOCATION 1
BURNER AIR SUPPLY

$T_s = 120^oF$ $\quad P_g = 29.18$
$T_{dry} = 120^oF$ $\quad P_s = +19.0"\ H_2O/30.58\ Hg$
$T_{wet} = 74^oF$

$$B_{ws} = \left(\frac{VP}{P_s}\right) \quad 100$$

where

VP = Vapor pressure H_2O @ stack cond.
P_s = Absolute pressure of stack gas

$$VP = SVP \left[(3.67 \times 10^{-4})\ (P_s)\ (T_d - T_w)\ \left(1 + \frac{T_w - 32}{1571}\right)\right]$$

where SVP = sat. H_2O vapor pressure, inches Hg at wet bulb temperature

$$VP = 0.8462 \left[(3.67 \times 10^{-4})\ (30.58)\ (120 - 74)\ \left(1 + \frac{74 - 32}{1571}\right)\right]$$

$$= 0.8462 - 0.5300$$

$$= 0.3161$$

$$B_{ws} = \left(\frac{0.3161}{30.58}\right) 100$$

$$= 1.03\%\ H_2O$$

PSYCHROMETRIC MOISTURE DETERMINATION -- LOCATION 2
PAINT BAKE OVEN OFF-GAS (INLET TO INCINERATOR)

T_{dry} = 282°F P_g = 29.18
T_{wet} = 118°F P_s = +7.4" H_2O/29.72 Hg

$$B_{ws} = \left(\frac{VP}{P_s}\right) \quad 100$$

where

VP = Vapor pressure H_2O @ stack cond.
P_s = Avsolute pressure of stack gas

$$VP = SVP - \left[(3.67 \times 10^{-4})\ (P_s)\ (T_d - T_w)\ \left(1 + \frac{T_w - 32}{1571}\right)\right]$$

$$VP = 3.259 - \left[(3.67 \times 10^{-4})\ (29.72)\ (282 - 118)\ \left(1 + \frac{118 - 32}{1571}\right)\right]$$

$$= 3.259 - 1.8867$$

$$= 1.3723$$

$$B_{ws} = \left(\frac{1.3723}{29.72}\right) 100$$

$$= 4.6\%\ H_2O$$

SAMPLE CALCULATION OF $\dot{Q}_j$ FOR INCINERATOR INLET STEAM

Determination of $\bar{C}_{p_j}$:

Component	$\bar{C}_{p_i}$ (cal/gm-mole-°C)	$X_i C_{p_i}(T_j - T_o)$
H_2O (liq)	18.0 }	41.82
H_2O (vap)	8.08 }	
CO_2	9.44	13.87
CO	6.98	0.47
N_2	6.98	604.96
O_2	7.10	141.20
		802.32

where $T_j = 411^oK$, $T_o = 298^oK$ and $T_{DP} = 304^oK$

$$\Delta h_{H_2O} = (10391)\,(0.043) = 446.81 \text{ cal/gm-mole}$$
$$\Delta h_r = (0.0006)\,(67410) + (376725)\,(195 \times 10^{-5})/1.8$$
$$= 40.45 + 40.81 = 81.26 \text{ cal/gm-mole}$$

$$\frac{\dot{Q}_j}{\dot{M}_j/(MW)_j} = 802.32 + 446.81 + 81.26 = 1330.39$$

$$\dot{Q}_j = \left(\frac{1330.39}{28.95}\right)(4.186 \times 10^3)\,(236.60)\,(60)$$

$$= 2731 \times 10^6 \text{ joules/hr}$$